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Sorbent enhanced water gas shift-rethinking carbon capture in IGCC

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Abstract

A dry sorbent pre-combustion CO_2 capture process to reduce carbon emissions and enhance the water gas shift (WGS) reaction in integrated gasification combined cycle (IGCC) power plants was investigated. The approach aims to eliminate the need for WGS reactors by maximizing the amount of H_2 through the removal of CO_2 and resulting changes in syngas chemical equilibrium. This concept can limit the energy penalty typically required of a WGS reactor by maintaining the higher temperature and pressure conditions present at a coal gasifier outlet and desired at the gas turbine. The sorbent needed for this process must have high reactive surface area, high CO_2 capacity, be able to tolerate required operating conditions (as high as 1000°C and 40 bar), and must have a long lifetime. A first principles approach was taken (thermodynamic and molecular modelling) to develop a list of candidate sorbents. Different approaches were taken to synthesize sorbents, including ultrasonic spray pyrolysis (USP), which led to materials with novel properties. These sorbents were characterized (*e.g.*, SEM, TEC, XRD), screened (*e.g.*, TGA) and ultimately tested in laboratory scale reactor systems at high temperatures and pressures.

A technoeconomic assessment based on laboratory results and a sorbent enhanced WGS (SEWGS) process, which takes advantage of the high heat of adsorption (ΔH_{ads}) of calcium oxide to generate turbine quality steam, is presented. The additional gross energy output resulting from the ΔH_{ads} helps offset the parasitic losses typically encountered for CO_2 capture. A slipstream of produced H_2 was used to regenerate the calcium sorbent in a 'regenerating boiler'; waste heat from this operation was recovered. Process designs were evaluated which increased the overall gross energy output of an IGCC by 40%, or from 737 MWe without CO_2 capture to 1,028 MWe with CO_2 capture for a fixed amount of coal. The energy produced from the ΔH_{ads} alone was estimated to contribute 429 MWe. In fact, the overall dynamics of produced energy are shifted from the majority of energy being produced from the gas turbine to a large fraction being produced by the CO_2 capture process itself.

In order to realize this SEWGS approach, scientific and engineering challenges must be met. Included are well designed adsorption and regeneration reactors which limit thermal shock, can efficiently remove the ΔH_{ads}, and can withstand H_2 combustion in the presence of a solid sorbent. Advancement in sorbent materials must also continue; sorbent replacement accounts for a large portion of O&M costs. Capital costs were projected to be high, but optimization should reduce this disadvantage. Even in light of these challenges, SEWGS with the 'regenerating boiler' concept can approach targeted increases in COE over IGCC without CO_2 capture of $81.30 per MWh through a rethinking of energy production in concert with CO_2 capture, rather than incremental cost reductions through evolutionary refinement of power generation augmented with CO_2 capture. The current optimal SEWGS case has a COE of $97.50 per MWh compared to $119.40 per MWh for IGCC with CO_2 capture.

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"Keywords: water gas shift, pre-combustion CO2 capture, sorbent, calcium carbonate, IGCC"

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1. Introduction

A conventional method to produce a high purity hydrogen gas stream from a coal gasification syngas involves a catalytic WGS reaction to convert CO and H_2O to H_2 and CO_2 and then remove the CO_2 in a separate unit, as shown in Fig 1. An alternative approach is to combine the WGS reaction with CO_2 capture in one unit, as shown in Fig 2. The combined process utilizes CO_2 adsorption in-situ on a solid sorbent bed and reduces or possibly even eliminates the need for a WGS catalyst.

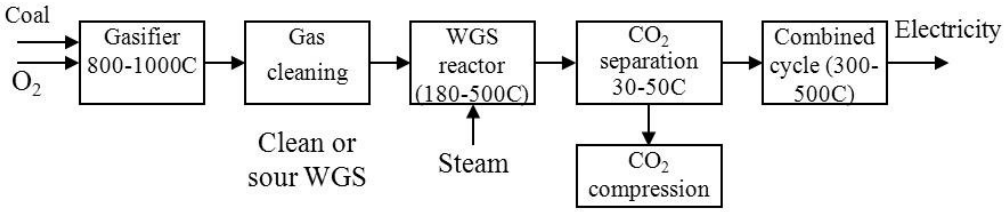

Fig 1. Flow Diagram of a Conventional IGCC Plant with CO_2 Capture

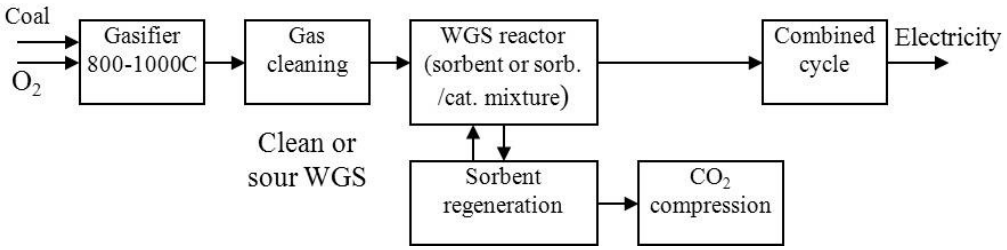

Fig 2. Flow Diagram of IGCC Plant with Sorption-enhanced WGS

Several studies have investigated solid sorbents for capturing CO_2. Harrison et al, [1, 2] Fan et al, [3, 4, 5] and Wang et al [6] studied sorption-enhanced WGS reaction using CaO sorbents. However, these sorbents (calcite, dolomite, huntite, etc.) tend to suffer considerable sintering problems after multiple adsorption/ regeneration cycles, especially at high regeneration temperature. Fan et al [3] showed that a precipitated calcium carbonate achieved 85%, 66.7%, and 45.5% calcium conversion in the 1st, 10th, and 100th cycle, respectively. A few synthetic sorbents such as lithium silicates (Li_4SiO_4) [7, 8] and sodium zirconates (Na_2ZrO_3) [8] were reported to have no appreciable loss of capacity or reactivity at high temperatures over several carbonation/regeneration cycles. A thermodynamic analysis showed that Na_2ZrO_3 had a higher H_2 yield and thermal stability than Li_4SiO_4. Both the CaO-based sorbents and the silicate or zirconate sorbents require regeneration at high temperatures to recover CO_2. For example, the Na_2ZrO_3 carbonation reaction reaches equilibrium at 790°C and P_{CO2} = 1 atm. $CaCO_3$ decomposes above 890°C if the CO_2 partial pressure is maintained at 1 atm. A vacuum condition is required to reduce the calcination temperature. Thus, the produced CO_2 is either under vacuum or at atmospheric pressure, implying a large energy penalty associated with CO_2 compression.

Lee et al [9, 10] reported CO_2 adsorption performances of Na_2O-promoted alumina and K_2CO_3 -promoted hydrotalcite. Both sorbents showed high selectivity of CO_2 chemisorption at 250-450°C. The byproduct CO_2 pressures reached 23 and 33 atm, respectively, for the two sorbents, during regeneration at 550°C with a super-heated, high-pressure steam. However, the overall CO_2 removal achieved during the adsorption/regeneration cycle was less than 60% of the CO_2 from the shift reaction. Further development of SEWGS requires advanced sorbents that can achieve a high level of CO_2 capture ($\geq$ 90%), demonstrate a high selectivity for CO_2 at high temperatures, and be able to produce a CO_2-rich gas at elevated pressure (>1 atm). Other important sorbent features include high sorption capacity, acceptable thermal stability and integrity, and minimal deactivation over long-term sorption cycles. Heat integration between the SEWGS and the IGCC plant should also be considered. A recent study [11] showed that a sorbent suitable for CO_2 adsorption at high temperatures tends to have a higher heat of adsorption. It is conceivable that the heat released during adsorption and the heat supply for the sorbent regeneration could be integrated with the IGCC plant to improve energy efficiency. This idea was explored through the technoeconomic assessment presented in this paper.

The WGS reaction is exothermic, which implies that equilibrium CO conversion increases with decreasing reaction temperature. In practice, however, both a high (300–500°C) and low temperature (180-300°C) shift catalyst are operated in series to convert most of the CO at high temperature and then achieve near complete CO conversion at a lower temperature. For CO_2 removal, the WGS gas is further cooled before entering an absorption or adsorption unit. The proposed CO_2 sorption-enhanced WGS has a number of significant advantages compared to conventional processes for H_2 production and CO_2 capture, including:

(1) Single Conversion Stage - Complete conversion of CO to CO_2 can be achieved at high temperatures (>400°C). As an example, this can be illustrated by an equilibrium analysis of the sorption-enhanced WGS using a CaO sorbent ($CO_2 + CaO = CaCO_3$, $\Delta H = -183$ *kJ / mol*), as illustrated in Fig 3. [1] At temperatures below 750°C, complete CO conversion is obtained at 25 atm total operating pressure compared to only 40% conversion without CO_2 removal. A sorption-enhanced WGS process employing an efficient sorbent could significantly reduce the amount of WGS catalyst necessary for a given CO conversion, or even eliminate the need for catalyst at high enough temperatures.

(2) High Conversion Rate - Achieving a higher CO conversion than limited by the WGS reaction equilibrium alone would reduce the WGS reaction steam requirements; this would subsequently increase plant efficiency and enhance operational flexibility of an IGCC system, including the CO_2 capture process.

(3) Operating at High Temperature - Removing CO_2 from the WGS gas at a temperature close to the gas turbine inlet temperature will eliminate the need for gas cooling/reheating thus improving thermal and economic performance of an IGCC plant.

(4) Production of Pure H_2 - The sorption-enhanced WGS permits direct production of a CO/CO_2-free H_2 gas stream at the feed gas pressure and eliminates the need for a separate CO_2 separation process. A high purity H_2 product gas can therefore be produced.

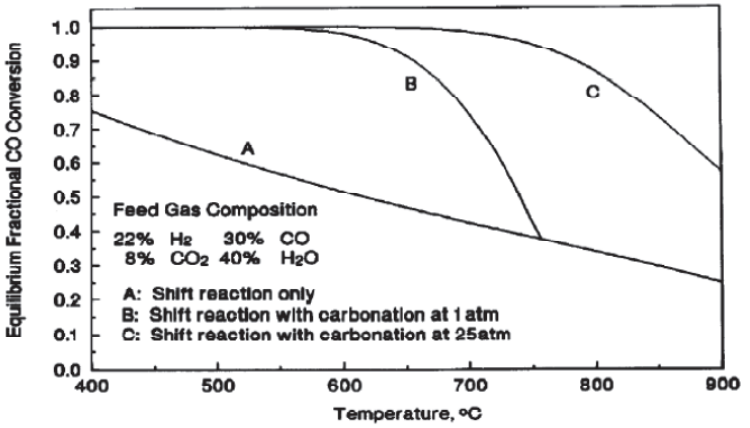

Fig 3. Thermodynamic Equilibrium Analysis: WGS and Sorption-enhanced WGS [1]

Efforts detailing the synthesis of sorbents [12, 13], computational simulations [14, 15], and experimental testing [16] have been presented elsewhere. The result of these studies pointed to a CaO based sorbent as having attractive thermodynamic properties [15], and results from laboratory tests indicate that engineered CaO sorbents can perform the desired WGS / CO_2 capture reaction including achieving the necessary 90% CO_2 capture [16]. Other groups have investigated the enhanced WGS approach using CaO sorbents [17]; this paper will focus on a possible process design and the economic assessment of that design especially as it relates to the DOE CO_2 Capture program goals. The objective of the technoeconomic assessment is to project the costs of a sorbent enhanced water-gas-shift (SEWGS) process using available thermodynamic and experimental data, reported elsewhere [15, 14, 13, 12, 16]. The analysis evaluates the capital costs associated with replacing or modifying an existing WGS reactor with a SEWGS process and will account for expected cost benefits associated with reduced WGS capital and operating costs, lower CO_2 compression costs (if applicable), and improved heat integration compared to standard cases presented in the DOE CO_2 Capture Report [18].

2. Methodology

The analysis utilized experimental data including key process data for newly-developed sorbents, the CO_2 capture performance and capacity at WGS-like conditions, and the performance degradation rates (to estimate sorbent lifetime). Additional data pertaining to water-gas-shift reaction efficiency in the presence of the CO_2 sorbents and regeneration purge gas requirements was also taken into account. Theoretical data associated with the heats of adsorption of the engineered sorbents and CO_2 partial pressure at optimal regeneration conditions was used in conjunction with experimental data to determine key parameters associated with heat management and CO_2 separation and compression costs.

This technoeconomic analysis was based on available experimental data, as described above. Because of the relatively immature nature of SEWGS technology, insufficient data exists to accurately project process design requirements or associated costs. Testing to date has been limited to laboratory-scale CO_2 adsorption and regeneration tests of powder sorbent materials. Until more detailed tests at larger scale are performed, it is not possible to accurately predict the actual performance and thermal properties of the sorbent (in the correct physical form) or the attrition rate over extended periods of continuous cycling; the latter information is important for determining sorbent life expectancy. Thus, various assumptions were required for these parameters in order to complete the current analysis, still taking results achieved to date into account.

A key objective of this analysis was the development of a conceptual design for a SEWGS reactor to be integrated into an IGCC plant. To determine the best possible reactor design, parameters associated with both the adsorption and desorption phases of the process were considered. Key adsorption parameters included required process conditions (with primary focus on temperature and pressure), sorbent CO_2 capacity, heat of reaction for a given sorbent, heat recovery scheme, optimization of gas-solid contacting, and gas pre-treatment requirements associated with impurities such as hydrogen sulfide. Key regeneration parameters included required process conditions (with primary focus on temperature and pressure), CO_2 partial pressure, potential purge gas type and requirements, heating requirements and source, and separation of CO_2 from the purge gas. Additional design considerations included the adsorption-regeneration cycling scheme and timing, overall heat management, and sorbent transport requirements; the latter focused on achieving optimal sorbent performance while minimizing attrition rate (assumed to increase as sorbent transport within the process increases).

The SEWGS process design includes estimates for the size and number of required reactors, for a plant of designated size (~550MWe), along with a reactor configuration for optimal gas-solid contacting. A process scheme is provided that enables continuous syngas treatment to be achieved for an operating plant. Based on estimated sorbent requirements, heat recovery needs (adsorption) and heat input needs (regeneration) were estimated along with a process design to achieve them. Capital costs for the designed process were estimated using Aspen Icarus software and other costing databases, and will be addressed in more detail below. Potential benefits of the SEWGS process, relative to a standard case, were also taken into accounted. These include potential cost benefits associated with the elimination or down-sizing of the existing WGS reactor, elimination or modification of an existing sulfur removal process, decreased CO_2 compression costs, any benefits associated with process heat management, and those associated with decreased hydrogen re-heating requirements upstream of a combined cycle combustion unit.

Costs were developed for two different SEWGS conceptual designs, however only one design is presented here. One approach was based on conservative assumptions and requires no appreciable advances in the current state of knowledge. This approach involves the use of fluidized bed reactors that serve as both adsorption and regeneration vessels for the SEWGS process, and resulted in unrealistic costs and unit operations and will not be discussed. The second design will be discussed in detail; this approach requires some technical challenges to be met and is based on a more aggressive engineering design. The assumptions made remain grounded in experimental and modeling results. This alternative approach assumes a moving sorbent bed with dedicated adsorber and regenerator reactors, with the latter resembling a boiler; and is termed a ‘regenerating boiler’.

This work presents a significant conceptual design for a full-scale integrated system using calcium based sorbents. The process involves the use of a dry sorbent material to adsorb CO_2 and then be regenerated, all at high temperatures and pressures. Furthermore, the adsorption process is highly exothermic resulting in the production of appreciable heat as CO_2 is removed from the syngas. The adsorption process must be maintained at a specific temperature and high pressure whereas the associated regeneration process requires that the sorbent be subjected to a even higher temperatures. These requirements carry with them the need for appropriate means in the process design to handle the extreme conditions, manage heat, and maintain appropriate levels of control.

3. Process Description

The general approach is shown schematically in Fig 4. A sorbent will be used in a high temperature, high pressure reactor to adsorb CO_2 while simultaneously shifting the WGS reaction towards H_2. The regenerating boiler will use dedicated adsorbers where CO_2 will react with the sorbent, CaO, to generate $CaCO_3$. The heat of reaction from the adsorbent will be used to generate turbine quality steam. The adsorber will conceptually be a fluidized bed integrated with a tube heat exchanger. Several nontrivial technical challenges exist to remove the heat of adsorption, namely:

- Minimizing the thermal shock of heating tubes containing cooling water with the high temperature gas,
- Removing the heat of adsorption, which may be restricted within a small region of the adsorption reactor.

The project team feels that these challenges can be met with creative engineering designs and accepted engineering concepts, and the approach warrants further study.

The spent sorbent will be continuously cycled from the adsorber to a dedicated reactor where the energy required to regenerate CaO will be supplied by the combustion of H_2 and O_2. The H_2 will come from a slipstream of shifted syngas while the O_2 will come from an additional (or, larger) air separation unit, which is required for the coal gasification step. The effluent from the regenerating boiler will be largely CO_2 and steam from the combustion of H_2 and O_2; some heat of combustion will be recovered at the regenerating boiler effluent to generate turbine quality steam. Recovering turbine quality steam from the regenerator effluent is an important component of this approach, but the downstream heat exchangers have not been optimized in this design. The regeneration step also presents technical hurdles, including combusting relatively pure H_2 and O_2 in the presence of a solid sorbent in a fluidized bed. Lessons can be learned from supercritical pulverized coal boilers, but unlike those design the solid must not only be captured but also recycled to the system. Moving the solid sorbent into and out of the reactor while maintaining the proper operating conditions (temperature as well as pressure) is a concern, and the high temperature of H_2 combustion necessitates the use of refractory lining materials inside of the reactor. Finally, the energy required to increase the gas stream and supply the heat necessary for calcination may not be evenly distributed throughout the regenerating boiler and creative solutions may again be necessary.

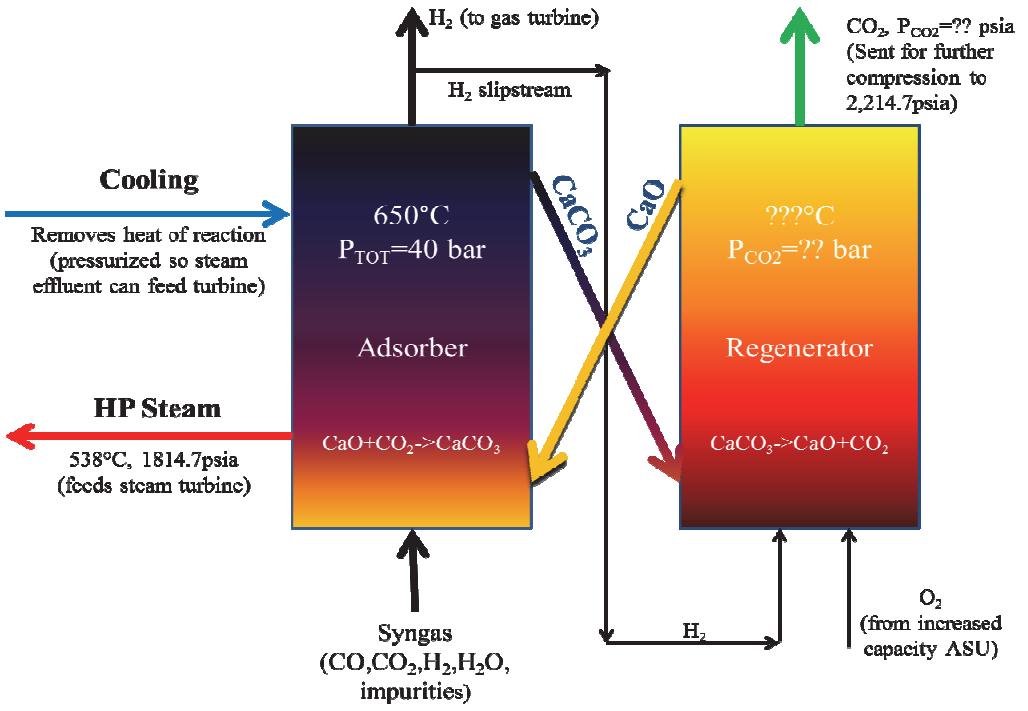

Fig 4. Simplified PFD for regenerating boiler approach to SEWGS, single pair of adsorber and regenerator reactors shown.

This approach can be summarized in the following bullets.

- Dedicated adsorbers and regenerators, so each reactor was specifically designed for a single purpose and the sorbent must be moved from adsorber to regenerator (and back again) at operating conditions. This limits the overall number of reactors required.
- Pressurized water (1,830 psi) was used to remove the heat of the adsorption such that the cooling media effluent was of a quality that can be used to feed a steam turbine (650°C, 40 bar). This required additional capital for increased steam turbine capacity but resulted in significantly enhanced power output.
- H_2 generated from the SEWGS adsorber was combusted with O_2 from the ASU to generate the heat for regeneration. This required additional capital for increased ASU capacity, resulted in additional parasitic losses from running the larger ASU, and reduced the power generated at the gas turbine due to the lost H_2. Some energy was recovered in the form of turbine quality steam.

Significant changes to base IGCC cases from the DOE CO_2 Capture Report [18] for energy added to the system, energy removed from the system, and capital cost implications are summarized below:

- *Energy Addition to the System*
  - Steam generated from CO_2 adsorption / heat removal from the adsorber
  - Steam generated the regenerator effluent
- *Energy Removed from the System (Parasitic Losses)*
  - H_2 used for regeneration and therefore lost to the gas turbine
  - Decreased steam production from the combined cycle (due to reduced H_2 at the gas turbine)
  - Increased duty for the ASU to supply O_2 for combustion with H_2 in the regenerating boiler
  - Pump duty to increase cooling tower water pressure to 1,830 psia for feeding adsorber cooling loops
  - Cooling CO_2 effluent prior to compression to pipeline pressure (note: some of the heat is recovered to produce turbine quality steam)
- *Capital Cost Implication*
  - Cost of the adsorber, regenerator, and conveyance system to move sorbent back and forth
  - Removing traditional WGS reactors lowers capital costs
  - Increased cost of the ASU to account for additional O_2 needed to combust with H_2 in the regenerating boiler
  - Larger steam turbine size because of added steam load
  - Smaller gas turbine size because of reduced H_2 load.

With these basic assumptions, five different scenarios were evaluated. For all cases, the adsorption parameters were as shown in Fig 4, but the following regeneration cases were evaluated, essentially differentiated by the operating conditions:

- Case A: T = 800°C, P_{CO2} = 0.8 bar
- Case B: T = 860°C, P_{CO2} = 1bar,
- Case C: T = 900°C, P_{CO2} = 1.9 bar,
- Case D: T = 1,075°C, P_{CO2} = 15 bar, and
- Case E: T = 1,165°C, P_{CO2} = 20 bar.

Aspen Plus V7.3 was used to represent the high pressure SEWGS process. Results obtained from the simulation were used to determine operating and design conditions, the process heat and mass balance (H&MB), equipment design, plant utilities requirements, and preliminary economic estimates. The simulation was generated using the following information:

- Inlet syngas composition and flow per stream 13, Exhibit 3-88 Case 6 Stream Table. [18]
- CO_2:CaO (sorbent) ratio of 0.3 lb/lb (demonstrated in laboratory scale experiments [16])
- Total H_2O:CO molar ratio of 1.5.
- Total H_2O (steam for regeneration):$CaCO_3$ molar ratio of 0.1.
- Density of fluidized sorbent after accounting for heat exchange tubes is 644 kg/m^3
- Base methods used:
  - Predictive Redlich-Kwong-Soave (PSRK), and
  - IDEAL property method that accommodates both Raoult's law and Henry's law. This method uses the:
    - Ideal activity coefficient model for the liquid phase ($\gamma = 1$).
    - Ideal gas equation of state PV = RT for the vapor phase.
    - Rackett model for liquid molar volume.

The model represents a single absorber / regenerator reactor pair using the steam/sorbent amounts calculated per the predefined ratios and associated ancillary equipment.

4. Process Economic Evaluation

The costs for the SEWGS system were estimated using a combination of Aspen Icarus and various scaling factors applied to the DOE NETL CO2 Capture Report. [18] Additional cost databases for commodity items, developed at URS through experience pricing capital projects, were referenced for materials such as pipe, manual valves, and instrumentation. Installation costs were estimated using an approximated construction schedule based on the scope of the project and assumed labor rates for the mid-Western U.S. These costs were combined with Case 6. [18] In many cases, scaling factors were developed and applied directly to line items from the DOE Report; this provided a more direct comparison. Included in these scaling factor assumptions were the increased steam turbine cost, the increased ASU cost, and decreased gas turbine cost. These scaling factors were developed based on the increased steam duty requirements, for example, and applied to capital costs from Case 6. This fidelity of this assumption could be improved but offered a tangible method to develop conceptual cost estimate costs. Other ancillary items included in the costs are as follows:

- Labor associated with installation,
- Construction consumables,
- Piping and ductwork, as appropriate based on operating conditions,
- All necessary valving and instrumentation and control, and
- Structural steel.

4.1. Equipment Sizing and Cost Sources

Equipment was sized as part of Aspen modeling. Large pieces of equipment are tabulated in Table 1. All equipment was carbon steel except the adsorber, regenerator, and the heat exchange equipment just downstream of the regenerator. These pieces of equipment, which will experience higher temperatures and pressures, will be lined with a refractory brick to insulate and dissipate heat and generally protect the raw metal surface. The O_2 compressor is only required for Cases D and E where regeneration is conducted at higher pressure than the O_2 that is generated from the air separation unit (ASU).

Table 1. Account of regenerating boiler SEWGS equipment, including design conditions and quantities.

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Adsorbers (including cyclones)	High temperature, pressure vessel	24 ft D x 35 ft H 139 MMscfd and 952 gpm 580 psia, 1,225°F	5	1
2	Regenerator (including cyclones, case specific)	Regenerating Boiler 14 ft D x 26 ft H	A: 9 MMscfd 20 psia, 1,495°F	5	1
			B: 12 MMscfd 20 psia, 1,605°F		
			C: 15 MMscfd 40 psia, 1,675°F		
			D: 27 MMscfd 395 psia, 1,990°F		
			E: 35 MMscfd 570 psia, 2155°F		
3	CO_2 Compressor (case specific)	Integrally geared, multi-stage centrifugal	A: 44 MMscfd @ 2,215 psia	5	1
			B: 44 MMscfd @ 2,215 psia		
			C: 42 MMscfd @ 2,215 psia		

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
			D: 35 MMscfd @ 2,215 psia		
			E: 27 MMscfd @ 2,215 psia		
4	O_2 Compressor (case specific)	Integrally geared, multi-stage centrifugal	D: 9 MMscfd @ 359 psia	5	1
			E: 29 MMscfd @ 524 psia		
5	Heat Exchanger I, CO_2 stream (case specific)	Shell and Tube Exchanger	A: 48 MMscfd, 38 gpm 27 MMBtu/hr	5	1
			B: 52 MMscfd, 51 gpm 37 MMBtu/hr		
			C: 55 MMscfd, 60 gpm 44 MMBtu/hr		
			D: 67 MMscfd, 111 gpm 80 MMBtu/hr		
			E: 74 MMscfd, 144 gpm 105 MMBtu/hr		
6	Heat Exchanger II, CO_2 stream (case specific)	Air cooling fans	A: 48 MM scfd 60 MMBtu/hr	5	1
			B: 52 MM scfd 71 MMBtu/hr		
			C: 55 MM scfd 82 MMBtu/hr		
			D: 67 MM scfd 118 MMBtu/hr		
			E: 74 MM scfd 139 MMBtu/hr		
7	Sorbent mover	Screw conveyor	3,543 lb/hr, 57 cfm	5	1
8	Sorbent hopper	Vessel	20 ft ID x 40 ft H	5	0

Also included were ancillary equipment, such as water pumps and all piping, valving and instrumentation. The air separation unit (ASU), steam turbine and gas combustion turbine were all sized to accommodate the case specific flows; these costs were adapted from the DOE Report as mentioned previously. The capacities were based on revised flows for these items and are presented in Table 2. Note that the gas feeding the steam turbine is increased by (a) the energy removed from the SEWGS adsorbing reactor and (b) the energy recovered downstream of the regenerator but lowered by (c) the reduced combined cycle steam produced at the gas combustion turbine. The gas turbine capacity is also reduced by the H_2 slipstream necessary to heat the regenerating reactor, and the ASU is increased by a similar amount to supply the O_2 needed for combustion.

Table 2. Sizing information for ASU, steam and gas combustion turbines.

Item	Case A	Case B	Case C	Case D	Case E
ASU (lbmol O_2 / hr generated)	12,028	12,714	13,190	15,421	16,809
Steam Turbine (lb/hr @ 1815 psia , 1000°F)	3,489,255	3,477,686	3,470,194	3,453,665	3,449,721
Gas Turbine (lbmol H_2 / hr)	32,887	31,518	30,545	26,115	23,413

4.2. Capital and Operating Costs

Details of the costs associated with the various cases modelled, in the format of the DOE CO_2 Capture Report [18] and including capital as well as operating costs, are presented elsewhere [16]. These details show that the line items associated with each case closely mimic Case 6 from the DOE CO_2 Capture Report, except the Selexol system and WGS reactors are excluded while the regenerating boiler SEWGS reactor system is added. Table 3 summarizes the cost of the SEWGS system specific items, including unit operations included in the base case IGCC system but adapted for the regenerating boiler SEWGS approach. Critical items and items that differ from Case 6 of the DOE Report for Cases A through E are presented in Table 3 (note that other items, such as insurance and finance charges, are dependent on total costs and therefore also differ, but are presented here only as rolled up costs). Costs are presented in 2007 dollars.

Table 3. Cost breakdown for regenerating boiler SEWGS system.

	2007 Dollars x 1,000,000				
	Case A	Case B	Case C	Case D	Case E
Adsorber (incl cyclones)	$52.07	$52.07	$52.07	$52.10	$52.22
O_2 Compressor				$11.04	$11.26
Regenerator (incl cyclones)	$46.72	$46.72	$49.76	$60.62	$68.59
Sorbent Mover and Hopper	$0.89	$0.90	$0.91	$0.95	$0.95
CO_2 Cooler (Stage I and II) and H_2O Separator	$4.94	$4.96	$4.96	$6.09	$6.73
CO_2 Compressor	$28.34	$28.34	$26.38	$14.29	$11.73
Gas Turbine	$91.58	$87.77	$85.06	$72.72	$65.20
ASU	$199.69	$211.08	$218.98	$256.02	$279.06
Steam Turbine	$134.71	$134.27	$133.98	$133.34	$133.19
Miscellaneous, SEWGS	$220.90	$220.90	$220.90	$220.90	$220.90
Total	$779.84	$786.99	$793.00	$828.08	$849.83

An itemized list of important capital and operating cost items is summarized below.

- Sorbent cost is assumed to be $1,032/ton. The sorbent cost has an estimated range of $185 - $2,670 per ton, depending on the feedstock being used and the synthesis method employed. Laboratory testing to date has shown a USP sorbent to perform the best; that sorbent cost is estimated to be $1,032/ton based on a 75:25 Ca:Al USP sorbent, an average USP energy use, and estimates of the raw feedstock and transportation. This analysis is described in more detail in Ref [16].
- Sorbent capacity of 0.3 $g_{CO2}/g_{sorbent}$ was assumed. This was a reasonable assumption based on the 12 cycle tests presented [16], but does require significant extrapolation.
- Reactor size was estimated based on the sorbent capacity, required heat exchange area, and gas residence times and developed using Aspen.
- Based on the two sorbent assumptions stated above, the initial sorbent load will be 3.544 Mlb per reactor. For 6 reactors (5 operating, 1 spare), this translates to 10,632 tons. The daily make up rate of sorbent is 121 tons/day assuming 4 effective change-outs per year per reactor, which is an extrapolation of results from the laboratory.
- A Claus unit is assumed.
- Sorbent disposal costs are assumed to be $75/ton for the effective make up rate of the sorbent (44,300 tons/year).

- CO_2 compression costs were estimated based on Case 6 of the DOE Report [18], adjusted to accommodate the different pressures of each specific regenerating boiler case. In other words, the cost from Case 6 was adjusted up or down depending on the flowrate and pressure requirements of the specific case being evaluated.
- Maintenance labor and maintenance materials was estimated based on factors applied to the total as spent capital costs (TASC).
- Preserved the 300 acre assumption per DOE Report for an IGCC with CO_2 capture. [18]
- An Engineering Fee of 9.27% was used.
- A 30% process and 30% project contingency was applied to the SEWGS capital equipment. All other process and project contingencies are per Case 6 of the DOE Report. [18]

4.3. De-Rating

The energy profile of a plant employing SEWGS via the regenerating boiler concept would be markedly different than the energy profile of a traditional IGCC. Less energy is generated via the gas turbine (as much as 35% less for Case E), but a significant amount of energy is produced by the steam generated from the heat of adsorption (approximately a factor of three when compared to a traditional IGCC). If more novel, cost-effective ways to regenerate the calcium based sorbent could be devised, this process becomes even more attractive. As it currently stands, a net energy output from the IGGC with SEWGS using the regenerating boiler can produce as much as 888 MWe for Case A. It should be noted that Case A has unattractive regeneration kinetics because of the low temperature used, but was economically evaluated in order to provide a larger range of operating conditions to consider. A summary of the energy generated and the parasitic loads for the five cases is presented in Table 4. Below is a summary of some of the important features of the de-rating analysis.

- The energy penalty associated with CO_2 compression is specific to the partial pressure of CO_2 in the regenerator effluent and was determined as a fraction of the energy penalty for compression for Case 6 from the DOE Report.
- Additional water usage is estimated based on the added cooling water requirements (case specific) and assuming a 50% recycle. The case specific water withdrawal, including the remainder of the IGCC, were determined to be as follows:
    - Case A: 6,595 gpm
    - Case B: 6,621 gpm
    - Case C: 6,640 gpm
    - Case D: 6,740 gpm
    - Case E: 6,806 gpm
- Parasitic losses associated with the additional capacity needed for the ASU were increased as fractions of the ASU energy penalty presented in Case 6 of the DOE Report.
- Miscellaneous balance of plant (BOP) equipment associated with the SEWGS not defined elsewhere account for 3.8 to 4.4 MW parasitic losses for Case A to Case E.

Table 4. Parasitic loads for regenerating boiler SEWGS compared to Case 5 and 6.

	Case 5	Case 6	SEWGS w/ Regen Boiler, Case A	SEWGS w/ Regen Boiler, Case B	SEWGS w/ Regen Boiler, Case C	SEWGS w/ Regen Boiler, Case D	SEWGS w/ Regen Boiler, Case E
POWER SUMMARY (Gross Power at Generator Terminals, kWe)							
Gas Turbine Power	464,000	464,000	423,632	406,000	393,472	336,400	301,600
Sweet Gas Expander Power	0	0	0	0	0	0	0
Steam Turbine Power	273,000	209,400	637,542	635,428	634,059	631,039	630,319
TOTAL POWER, kWe	**737,000**	**673,400**	**1,061,174**	**1,041,428**	**1,027,531**	**967,439**	**931,919**
AUXILIARY LOAD SUMMARY, kWe							
Coal Handling	440	460	460	460	460	460	460
Coal Milling	2,040	2,170	2,170	2,170	2,170	2,170	2,170
Slag Handling	520	550	550	550	550	550	550
Air Separation Unit Auxiliaries	1,000	1,000	1,151	1,217	1,262	1,476	1,608
Air Separation Unit Main Air Compressor	45,190	59,740	68,755	72,676	75,397	88,151	96,085
Oxygen Compressor	8,890	9,460	9,460	9,460	9,460	11,614	13,421
Nitrogen Compressor	29,850	32,910	30,047	28,796	27,908	23,860	21,392
CO_2 Compressor		30,210	30,210	30,210	22,359	6,485	3,821
Boiler Feedwater Pumps	4,500	3,500	5,000	5,020	5,034	5,111	5,161
Condensate Pump	230	280	400	402	403	409	413
Quench Water Pump		610	872	875	877	891	899
Syngas Recycle Compressor	680	790	790	790	790	790	790
Circulating Water Pump	3,400	4,370	6,243	6,243	6,243	6,243	6,243
Ground Water Pumps	370	510	729	731	734	745	752
Cooling Tower Fans	1,760	2,260	3,229	3,242	3,251	3,300	3,332
Scrubber Pumps	770	360	360	360	360	360	360
Acid Gas Removal (Selexol or SEWGS)	620	18,650	3,767	3,855	3,721	4,233	4,420
Gas Turbine Auxiliaries	1,000	1,000	913	875	848	725	650
Steam Turbine Auxiliaries	100	100	304	303	303	301	301
Claus Plant/TGTU Auxiliaries	250	250	250	250	250	250	250
Claus Plant TG Recycle Compressor	890	1,830	1,830	1,830	1,830	1,830	1,830
Miscellaneous Balance of Plant[2]	3,000	3,000	3,000	3,000	3,000	3,000	3,000
Transformer Losses	2,520	2,530	2,530	2,530	2,530	2,530	2,530
TOTAL AUXILIARIES, kWe	**108,020**	**176,540**	**173,020**	**175,845**	**169,740**	**165,483**	**170,438**
NET POWER, kWe	**628,980**	**496,860**	**888,154**	**865,583**	**857,791**	**801,956**	**761,480**
Net Plant Efficiency, %(HHV)	**42.1**	**31.2**	**55.8**	**54.4**	**53.9**	**50.4**	**47.8**

4.4. Economic Analysis and Results

The Cost of Electricity (COE) was calculated for the IGCC cases with regenerating boiler SEWGS and compared to the IGCC without CO_2 capture (Case 5) and with Selexol (Case 6) and is presented in Table 5. The COE was developed per DOE guidelines and includes the following assumptions:

- Capital charge factor of 12.43% (High Risk, IOU, Five Years).
- Fixed and variable operating costs detailed in Ref [16].
- Capacity factor of 80%.
- Coal usage and CO_2 TS&M costs that mimic Case 6.
- Levelization factor of 1.268%.

Fig 5 shows the capital equipment for the regenerating boiler cases, with different categories of equipment represented by different colored bars. The COE overlays the bar chart and corresponds to the secondary y-axis. Starting at the bottom of the chart, the equipment categorized as 'all other equipment' was the same for all cases, and can be considered all IGCC capital costs that are invariant and also common to Case 5 and Case 6 from the DOE Report. Miscellaneous equipment is specific to the SEWGS, but also does not change from case to case. Beginning with the ASU, capital costs are different for each regenerating boiler SEWGS case and, as regeneration temperature increases from Case A to Case E, the size/cost of the ASU also increases. Conversely, the capital costs for the gas turbine are seen to decrease as a larger H_2 slipstream is necessary to heat the regenerator to higher temperatures. The steam turbine costs are largely the same throughout the five cases. If taken together, the CO_2 and O_2 compressors costs are approximately the same, meaning that the capital savings realized by regenerating at a higher pressure was offset because a compressor was needed to increase the O_2 pressure feeding the regenerators. Not surprisingly, the cost of the regenerators themselves also begins to climb as the regeneration conditions become more extreme.

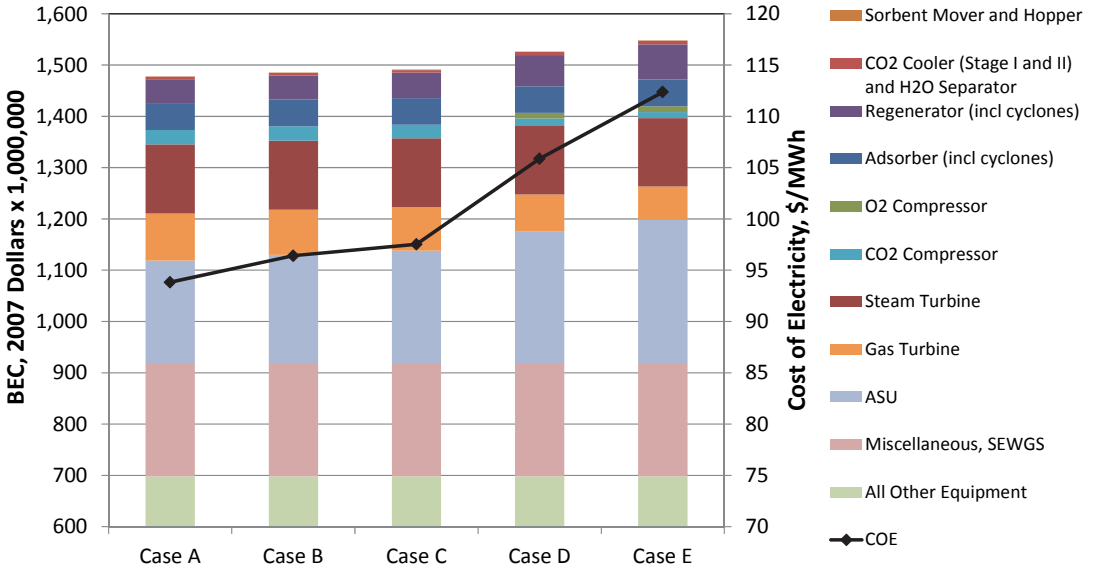

Fig 5. Bar chart for Cases A through D: Capital costs and COE.

Table 5. Cost of electricity for regenerating boiler SEWGS

	Case 5	Case 6	SEWGS w/ Regen Boiler, Case A	SEWGS w/ Regen Boiler, Case B	SEWGS w/ Regen Boiler, Case C	SEWGS w/ Regen Boiler, Case D	SEWGS w/ Regen Boiler, Case E
TOC =	$ 1,708,524,000	$ 1,939,878,000	$ 2,565,070,630	$ 2,574,939,704	$ 2,585,280,261	$ 2,649,993,280	$ 2,686,811,811
CCF (High Risk IOU, 5 yrs) =	12.43%	12.43%	12.43%	12.43%	12.43%	12.43%	12.43%
OC_{FIX} =	$ 53,508,812	$ 58,210,058	$ 79,367,630	$ 79,641,584	$ 79,928,650	$ 81,725,222	$ 82,747,354
OC_{VAR} =	$ 34,165,649	$ 34,629,764	$ 88,706,884	$ 88,884,911	$ 89,068,511	$ 90,210,994	$ 90,864,282
CF =	80%	80%	80%	80%	80%	80%	80%
MW =	629	497	888	866	858	802	761
Coal Usage, Tons / Day =	5240	5583	5583	5583	5583	5583	5583
	COE CALCULATION						
Capital	48.2	69.2	51.2	52.8	53.5	58.6	62.6
Fixed OCs	12.1	16.7	12.8	13.1	13.3	14.5	15.5
Variable OCs	7.8	9.9	14.3	14.7	14.8	16.1	17.0
Fuel Cost	13.3	17.9	10.0	10.3	10.4	11.1	11.7
CO2 TS&M	0.0	5.6	5.6	5.6	5.6	5.6	5.6
COE, mills/kWh or $/MWh	81.3	119.4	93.8	96.4	97.5	105.9	112.4
LCOE (LF = 1.268%)	1.03	1.51	1.19	1.22	1.24	1.34	1.42

The COE is seen to moderately increase from Case A to Case C, and then – as the regenerator conditions become more extreme – significantly increase for Cases D and E. The capital costs do not increase dramatically for Cases D and E, rather the increase in COE is the result of an energy penalty from the increase in regenerator temperature, which is only partially recovered as steam. The COE also generally suffers because of large fixed and variable operating costs; some of these costs are from the sorbent while other line items were tied to the TOC/TASC and increase as those values increases. It is reasonable to assume that the material maintenance costs of operating the IGCC with SEWGS would not be significantly higher than IGCC with Selexol, but since this value is tied to the TOC it increases by 50%. However, even given some of these high costs, the COE is appreciably lower than IGCC with Selexol and is approaching DOE goals. More than half of the DOE goal of a 10% increase in COE for IGCC with CO_2 capture is taken in transmission, storage and monitoring costs; meaning only ~$5/MWh is available for CO_2 capture. The project team believes that the only way to successfully approach this goal is to rethink and redesign the way energy is produced in concert with carbon capture. An increase in capital costs seems unavoidable, so increasing the energy output from the plant without an increase in coal usage is one approach to maintaining a modest increase in COE.

Regarding the sensitivity of the regenerating boiler concept to different variables, investigating the five different cases effectively brackets many of the important variables that can impact the economics of the SEWGS process. Capital costs and COE increase as the regenerating temperature and pressure increase. Because of the slower reaction kinetics involved, one cannot simply choose the lowest temperature (and cost) option, Case A. Regenerating at 900°C (Case C) is likely the most attractive option when considering both economic and technical drivers. Regeneration has been successfully conducted at 900°C in the laboratory.

It is not practical to reduce the number of reactors any further, based on the sorbent proposed for use and its properties, the amount of heat necessary to move through the system, and residence times required. Therefore, it is difficult to quantify significant decreases in the capital costs, though some can certainly be realized through engineering advances and application of the technology multiple times. Advances in sorbent properties, such as decreased cost or increased lifetime, can quantifiable lower the COE.

For example, if the sorbent cost was reduced from $1,032/ton to $516/ton (a 50% decrease but still within the range of projected prices presented in Ref [16]), the COE for Case C would be reduced from $97.50 to $94.20 per MWh, a 3.3% decrease. Similarly, an increase in lifetime resulting in a reduction from four change-outs per year to one change-out per year reduces the COE to $92.80, a 4.8% decrease. Applying both of these changes reduces the COE to $91.90, an overall reduction of 5.7% and only a 13% increase in COE over IGCC without CO_2 capture technology (Case 5, $81.30/MWh,).

5. Conclusions

Results of a technoeconomic analysis investigating a SEWGS process design that utilizes the heat generated during the CO_2 adsorption process and allows for optimized design of reactors used for adsorption and sorbent regeneration was presented. The 'regenerating boiler' approach uses a dry engineered sorbent in a high temperature, high pressure reactor to adsorb CO_2 while simultaneously shifting the WGS reaction towards H_2. The approach includes dedicated adsorber vessels where CO_2 reacts with a CaO-based sorbent to generate $CaCO_3$. The adsorption reactors were designed to efficiently capture the heat generated from the CO_2 adsorption process using a pressurized water media resulting in the generation of turbine quality steam. The adsorber is conceptually a fluidized bed integrated with a tube heat exchanger. Spent sorbent is continuously cycled from the adsorber to a dedicated regeneration reactor where the energy required to regenerate the CaO material is supplied by the combustion of H_2 and O_2. The H_2 will come from a slipstream of shifted syngas while the O_2 will come from an additional (or, larger) air separation unit, which is required for the coal gasification step. The effluent from the regenerating boiler is largely CO_2 and steam from the combustion of H_2 and O_2; some heat of combustion will be recovered to generate turbine quality steam.

The regenerating boiler approach relies on the ability to move dry sorbent from dedicated adsorbers to dedicated regeneration vessels operating at high temperatures and pressures. Although this provides the benefit of allowing those reactors to be designed for their specific task, the need to transport high volumes of material between vessels operating at extreme conditions will provide a technical challenge to the development of this technology. Additional challenges associated with sorbent attrition rates and lifetime must be addressed in tests conducted at larger scale.

The costs for the regenerating boiler SEWGS system were estimated using Aspen Icarus and additional cost databases. The costs were combined with Case 6 from the DOE NETL CO_2 Capture report as appropriate. [18] Scaling factors were developed based on flows and other operating conditions and applied to line items from Case 6

of the DOE Report, providing a more apples-to-apples comparison. Specific assumptions made for this analysis include:

- Assumed sorbent cost was $1,032/ton.
- Sorbent capacity of 0.3 $g_{CO2}/g_{sorbent}$ was assumed.
- Reactor size was estimated based on the sorbent capacity, required heat exchange area, and gas residence times.
- The process will consist of five operating reactor pairs (i.e., adsorption and regeneration) and one spare set.
- Based on the assumptions stated above, the initial sorbent load will be 3.544 Mlb per reactor (10,632 tons for 6 reactors).
- The daily sorbent make-up rate is 121 tons/day assuming 4 effective change-outs per year per reactor.
- A Claus unit was assumed.
- Sorbent disposal costs are assumed to be $75/ton for the effective make up rate of the sorbent (44,300 tons/year).
- CO_2 compression costs were estimated based on Case 6 of the DOE Report, adjusted to accommodate the different pressures of each specific regenerating boiler case.

Five different regeneration scenarios were investigated in an attempt to determine optimal conditions (*i.e.*, temperature, pressure) for the process. Here, regeneration temperature was varied from 800°-1165°C and pressure was varied from 0.8-20 bar. Conditions for CO_2 adsorption (*i.e.*, 650°C, 40 bar) were the same for each case. Energy trade-off associated with varying process conditions include faster regeneration kinetics at higher temperature (*i.e.*, more energy required). In addition, equilibrium favors sorbent regeneration as the CO_2 partial pressure decreases (thus, requiring higher regeneration temperatures as the system pressure is increased); thus a trade-off between higher compression costs (at lower operating pressure) and higher heat requirements (higher temperature needed at higher operating pressure) must be considered in order to determine optimal SEWGS process conditions.

When considering the different cases, the process oxygen requirements and electricity generation from both the steam and gas turbines varied depending on the process regeneration conditions as presented in Table 2, due to the interdependencies of the various (major) components. The gas feeding the steam turbine is increased by both the energy removed from the SEWGS adsorbing reactor and the energy recovered downstream of the regenerator but lowered by the reduced combined cycle steam produced at the gas combustion turbine. The gas turbine capacity is also reduced by the H_2 slipstream necessary to heat the regenerating reactor, and the ASU requirements are increased by a similar amount to supply the O_2 needed for combustion.

One dramatic finding of this analysis was associated with the energy profile of an IGCC plant employing SEWGS via the regenerating boiler approach. Less energy would be generated via the gas turbine with SEWGS, due to parasitic losses, but a significant amount of energy is produced by the steam generated from the heat of adsorption (approximately a factor of three when compared to a traditional IGCC). If more novel, cost-effective ways to regenerate the calcium based sorbent could be devised, this process becomes even more attractive. Based on this analysis, a net energy output from the IGGC with SEWGS using the regenerating boiler can produce as much as 888 MWe (Case A in analysis), thus substantially higher than ultimately produced in Case 6.

A summary of the energy generated and the parasitic loads for the five cases is presented in Table 4. Some of the important features of the de-rating analysis included:

- The energy penalty associated with CO_2 compression is specific to the partial pressure of CO_2 in the regenerator effluent and was determined as a fraction of the energy penalty for compression for Case 6 from the DOE Report.
- Additional water usage was estimated based on the added cooling water requirements (case specific) and assumed a 50% recycle. The case specific water withdrawal, including the remainder of the IGCC, ranged from 6,595 to 6,806 gpm for the cases evaluated.
- Parasitic losses associated with the additional capacity needed for the ASU were increased as fractions of the ASU energy penalty presented in Case 6 of the DOE Report.
- Miscellaneous balance of plant (BOP) equipment associated with the SEWGS not defined elsewhere account for 3.8 to 4.4 MW parasitic losses for Case A to Case E.

For the Regenerating Boiler approach, the TOC for an IGCC plant with a Shell gasifier and SEWGS ranged from $2.565M - $2.687M for the different cases evaluated. This compared to a TOC value of $1,940M for a similar IGCC with Selexol for CO_2 capture (i.e., Case 6). The COE for the regenerating boiler approach ranged from $93.8 - $112.4/MWh for the different cases evaluated, comparing favorably to COE value of $119.4.MWh for Case 6.

This shows appreciable improvement over the baseline case and indicates the appreciable benefit provided by the additional power output provided by the regenerating boiler approach.

Projected capital costs and COE for the regenerating boiler SEWGS process increase as the regeneration temperature and pressure increase. Because of the slower reaction kinetics involved, one cannot simply choose the lowest temperature (and cost) option. The projected COE was seen to moderately increase as the SEWGS regeneration operating conditions become more extreme, significantly increasing as the regeneration temperature increased over 1000°C (Cases D and E in the analysis). The capital costs did not increase dramatically for Cases D and E, rather the increase in COE was the result of an energy penalty from the increase in regenerator temperature, which is only partially recovered as steam. Regenerating at 900°C (Case C in the analysis) is likely the most attractive option when considering both economic and technical drivers. Regeneration has been successfully conducted at 900°C in the laboratory. More novel ideas regarding regenerator design and heat integration/management could lead to more significant cost savings. Improvement in sorbent design, resulting in higher CO_2 capacities or enhanced lifetimes, could offer opportunity to further lower the COE for this process.

An initial motivation for exploring the feasibility of SEWGS technology was the premise that the ability to conduct CO_2 adsorption and regeneration reactions at elevated pressures would result in an appreciable energy savings associated with CO_2 compression. Based on the results presented here, it appears that, at least for the SEWGS system, regenerating such that the effluent CO_2 gas stream is at high pressure, and thus more amenable to compression and sequestration / reuse, is not of primary importance. In fact, the energy penalty associated with compression (~30 MWe) is very small relative to the additional energy from the heat of adsorption (> 400 MWe). In its current design, the regenerating boiler only recovers a small fraction of the energy spent on regeneration, and enhanced heat integration can increase that fraction and potentially mitigate additional energy losses associated with CO_2 compression.

The regenerating boiler concept is a new approach to power generation with CO_2 capture. More energy is generated from steam than from the gas combustion turbine, and much of that steam is the result of the CO_2 adsorption process. The flow of energy through the plant is appreciably different, and the efficiencies are significantly higher than IGCC without CO_2 capture (56% for Case A, 48% for Case E). This fact tantalizes the opportunities that process optimization, design improvements, and more sophisticated heat integration may yield an even higher net energy output. Energy is being harnessed from a series of CO_2 capture unit operations that typically result in an energy loss. In short, unlike more established methods for energy generation with carbon capture, this novel approach has opportunity for improvement and a great deal of usable energy, some of which is already recovered.

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References

[1] Harrison et al, "Simultaneous Shift Reaction and Carbon Dioxide Separation for the Direct Production of Hydrogen," *Chemical Engineering Science 49(248),* pp. 5875-5883, 1994.

[2] Harrison et al, "Hydrogen Production Using Sorption-Enhanced Reaction," *Ind. Eng. Chem. Res. 40,* pp. 5102-5109, 2001.

[3] Gupta et al, "Enhanced Hydrogen Production Integrated with CO2 Separation in a Single-Stage Reactor," *Annual Technical Progress Report, DE-FC26-03NT41853,* 2004.

[4] Iyer et al, "Enhanced Hydrogen Production Integrated with CO2 Separation in a Single-Stage Reactor," *Annual Technical Progress Report, DE-FC26-03NT41853,* 2005.

[5] Iyer et al, "Enhanced Hydrogen Production Integrated with CO2 Separation in a Single-Stage Reactor," *Annual Technical Progress Report, DE-FC26-03NT41853,* 2006.

[6] Wang et al, "Thermodynamic Equilibrium Analysis of Hydrogen Production by Coal Based on Coal/CaO/H2O Gasification System," *Inter. J. Hyd. Ener. 31,* pp. 945-952, 2006.

[7] Kwang et al, "High Temperature CO2 Absorbents for Hydrogen Production through Sorption Enhanced Steam Reforming," *Nordic Hydrogen Seminar, Oslo, Norway.,* 2006.

[8] Bretado et al, "Thermodynamic Analysis for the Production of Hydrogen through Sorption Enhanced Water Gas Shift (SEWGS)," *International Journal of Chemical Reactor Engineering, 6,* pp. 1-16, 2008.

[9] Lee et al, "Reversible Chemisorption of Carbon Dioxide: Simultaneous Production of Fuel-Cell Grade H2 and Compressed CO2 from Synthesis Gas," *Adsorption, 13,* p. 385–397, 2007.

[10] Lee et al, "Performance of Na2O Promoted Alumina as CO2 Chemisorbent in Sorption-Enhanced Reaction Process for Simultaneous Production of Fuel-Cell Grade H2 and Compressed CO2 from Synthesis Gas," *Journal of Power Sources, 176,* pp. 312-319, 2008.

[11] Ciferno et al, "Analysis of Hot/Warm CO2 Capture for IGCC Processes," *AIChE Spring National Meeting, New Orleans, LA.,* 2008.

[12] Sayyah et al, "Mechanical Activation of CaO-Based Adsorbents for CO2 Capture," *ChemSusChem, 6,* p. 193–198, 2013.

[13] Sayyah et al, "CaO-based sorbents for CO2 capture prepared by ultrasonic spray pyrolysis," vol. 3, no. 43; 19872-19875, 2013.

[14] Zhang et al, "Sintering of calcium oxide (CaO) during CO2 chemisorption: a reactive molecular dynamics study," vol. 14, no. 47; 166333-16643, 2012.

[15] Lu et al, "CO2 sorbents for a sorption-enhanced water–gas-shift process in IGCC plants: A thermodynamic analysis and process simulation study," vol. 38, no. 16; 6663-6672, 2013.

[16] Richardson et al, "Evaluation of Dry Sorbent Injection Technology for Pre-Combustion CO2 Capture; Final Report," DOE, 2013.

[17] Chen et al, "Incorporating IGCC and CaO sorption-enhanced process for power generation with CO2 capture," vol. 95, no. 0; 285-294, 2012.

[18] "Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 2, 1397," *Department of Energy (DOE) National Energy Technology Laboratory (NETL),* 2010.